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WATER RECLAMATION STUDY PROGRAM

George Tchobanoglous, et al

Stanford University
Stanford, California

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PROGRESS REPORT

WATER RECLAMATION STUDY PROGRAM

DEMONSTRATION PROJECT GRANT

NO. WPD 21-05

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Water Quality Control Laboratory

Department of Civil Engineering

Stanford University

Stanford, California

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Covering the Period

August 16, 1966 to September 15, 1967

OCTOBER 1967

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SUMMARY

During the past five years, Stanford University has been operating a demonstration plant for the removal of nutrients (nitrogen and phosphorus) from treated sewage effluent. The demonstration plant, during this report period, was comprised of a package plant activated sludge sewage treatment unit, a pressure filter, and an ion exchange column. The pilot plant flow rate varies between 8.5 to 10.0 gpm. In operation, activated sludge effluent is passed through the filter unit, which contains sand and anthrafilt, and then through the anion exchange unit until about 190 bed volumes (BV) have been processed. Regeneration of the ion exchange resin was accomplished using sodium chloride and restoration of the resin was accomplished using an acid-methanol treatment. Effluent from the ion-exchange column is discharged to a lined observation pond.

Using a strong base anion exchange resin (Duolite A-1020) it has been possible to reduce the total phosphorus in the sewage effluent by an average of 88.2 percent (average effluent concentration, 0.60 mg/l) nitrate nitrogen by 82.3 percent (average effluent concentration, 1.1 mg/l). These removals closely conform to those predicted on the basis of laboratory studies. The estimated cost for nutrient removal (nitrogen and phosphorus) using the ion exchange process is about 15.00 to 17.43 cents per 1000 gallons. Of this total, chemical costs will vary between 6.40 and 7.28 cents per 1000 gallons.

In addition to the study of the ion-exchange process and the development of cost data for nutrient removal, other investigations undertaken during the period covered have included 1) the study of the removal of the various forms of phosphorus and the implication of this phenomena, 2) the study of parameters which control the efficiency of the filtration process as applied to the filtration of secondary effluent, 3) virus removal studies, and 4) the development of mathematical models which can be used to describe each of the units in the pilot plant process flowsheet.

WATER RECLAMATION STUDY PROGRAM
(Progress Report)

INTRODUCTION

Over the past five years the Stanford University Demonstration Plant has been used to make extensive evaluations of several tertiary treatment schemes for wastewater renovation. Among these were diatomite filtration, chemical precipitation, sand filtration, and ion exchange. These systems have been discussed in earlier progress reports dated August 1964, October 1965. and August 1966. During the past two years the demonstration plant as shown in Figure 1 has consisted of an activated sludge unit followed by a sand filter and a strong base anion exchange column. The anion exchange and sand filtration processes have obvious benefits applicable to the renovation of treated sewage effluents. The anion exchange unit produces a large improvement in the quality of the wastewater but is highly susceptible to physical clogging and chemical and physical degradation of the resin exchange capacity. While the filtration process itself does not produce the water quality improvement which is normally required, investigation of the parameters which govern its performance is of tremendous importance due to its applicability to many tertiary and wastewater renovation systems.

During the period covered by the last progress report (August 1966) (1) the sand filter and anion exchange column were redesigned to alleviate inherent difficulties in the effectiveness and operation of the units. Tests on the old and new units substantiated the need for their redesign. During the period covered by the present report efforts were made to optimize the efficiency of all processes of the demonstration plant including the activated sludge, sand filter, and ion exchange. The ultimate purpose of this plant process optimization was to decrease operating costs while maintaining desired removal efficiencies.

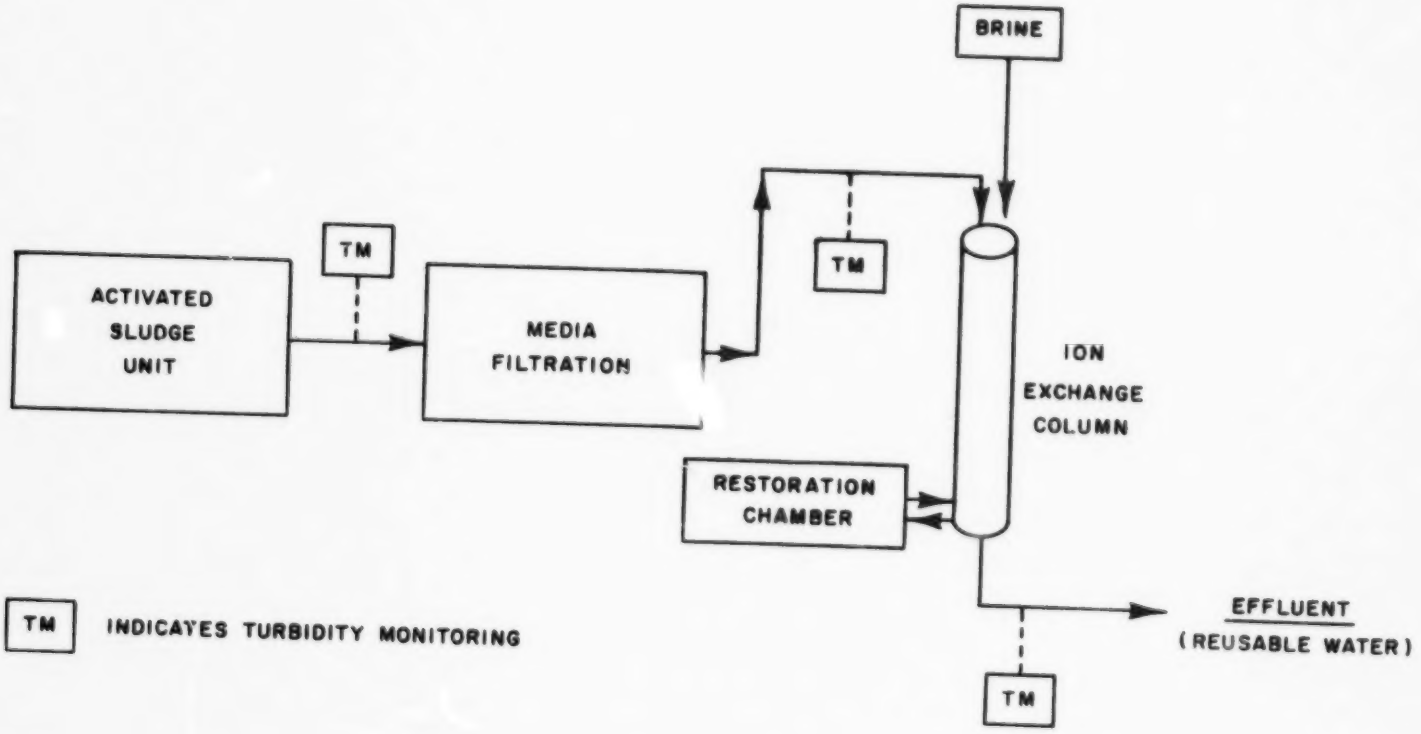


FIGURE 1
PILOT PLANT FLOW SHEET

PROGRAM OBJECTIVES

The overal objective of the demonstration plant operation is to investigate and define the potential benefits, particular advantages and disadvantages, and general applicability of the ion exchange and sand filtration processes for reclamation of treated wastewater. The current period of operation has been devoted to the optimization of the cost of existing processes. Two of the problems studies were the design and operation of filters for more effective utilization of the ion-exchange process and the development of satisfactory resin restoration techniques for minimizing the operating costs of the ion-exchange process.

PILOT PLANT DESCRIPTION

The process flow sheet of the 10 gpm demonstration plant which was used during the current study period is shown in Figure 1.

Equipment

Brief descriptions of each of the plant components are presented below.

Activated Sludge Unit. The activated sludge unit is pre-assembled "Rapid-Bloc" type unit which was provided by the Chicago Pump Company, Hydrodynamics Division, FMC Corporation. The units consists of an aeration chamber, a settling tank, and aerobic digestion compartment. It is operated as a secondary treatment unit, with an overall (aeration tank plus settling tank) theoretical detention time of 5.9 hours at the 10 gpm flow rate. The capacity of the aerobic digestion compartment is 295 cubic feet. A comminutor is included, although the unit normally receives settled primary sewage (effluent from the Palo Alto Sewage Treatment Plant primary settling tanks) as the influent liquid.

Sand Filter. The sand filter was fabricated and supplied by the Eimco Filter Corporation of Salt Lake City, Utah. The filter can be operated under either pressure of gravity conditions. The filtration media is supported by 21

plastic-and-nylon "Spraying Fulljet" media supports which provide a very uniform backwash water distribution. Air agitation of the media for 10-20 minutes is done immediately prior to backwashing the filter. The air agitation serves the same purpose as a water "surface wash," and has thus far prevented the formation of mudballs, or the accumulation of foreign matter in the filter. Viewing ports are provided so that the condition of the sand can be readily observed. The filter is 3 feet in diameter and operates at a flow rate of about 1.4 gallons per minute per square foot of surface area. The filter is backwashed at a flow rate of 125 gpm, producing a rise rate of 28.5 inches per minute. The filter media presently in use are sand and anthrafilt. The characteristics of the sand include an effective size of 0.52 mm and a uniformity coefficient of 1.48. The characteristics of the anthrafilt include an effective size of 1.47 mm and a uniformity coefficient of 1.70. It should be noted that the characteristics of the anthrafilt will vary with usage.

Ion Exchange Column. The ion-exchange column was fabricated and provided by the California Filter Company, of Burlingame, California. Incorporated into the ion-exchange column currently in use are many features designed to eliminate difficulties and uncertainties which had been encountered with the original column which was replaced early in 1966. Major improvements which were made included 1) elimination of the sand-gravel support bed through the use of specially designed underdrain appurtenances, 2) changes in the resin depth-to-area ratio and the resultant reduction of the volume of resin required (the redesign, by assuring better distribution of influent water, allowed the use of a flow rate of 26 bed volumes per minute instead of the 10 BV per minute rate used in the original column), 3) provision of adequate freeboard for resin expansion (up to 113 percent) during backwashing and regeneration, and 4) incorporation of a restoration chamber as an inherent part of the ion-exchange equipment.

Ion Exchange Resin. Duolite A-102D, a strong base anion exchange resin with quaternary ammonium functional groups is the type of ion-exchange resin used. This resin is supplied in porous bead form and is operated on the

chloride cycle (brine is used for regeneration). The resin was supplied by the Western Division of the Diamond Alkali Company, Redwood City, California. Duolite A-102D was selected as the ion-exchange resin on the basis of laboratory work in which a number of types and varieties of ion-exchange resins were tested.

Observation Pond

The observation pond, which has a maximum depth of four and one half feet and overall capacity of 250,000 gallons, was used to received the treated effluent from the ion-exchange unit. The effect of the nutrient removals resulting from the filtration-ion exchange treatment can be evaluated under natural conditions of sunlight and temperature, and the elimination of algal blooms and effect on algae growth in general can be demonstrated. The pond also provides information relative to bacterial and virus populations in the finished water.

Some difficulties had previously been experienced with seepage, and it was also felt that nutrients may have been "leached" from the underlying soil. To eliminate these uncertainties it was decided to line the pond, provided an effective and economical liner material could be obtained. A liner was obtained on a loan basis from the Griffolyn Company, Inc., of Houston, Texas, and it has proved to be extremely durable and resistant to punctures, tears, and light exposure problems.

PILOT PLANT OPERATION

The operational sequence of the demonstration plant is indicated by the arrows shown in Figure 1. Activated sludge effluent is passed through a dual media filtration unit and then through the anion exchange column until approximately 190 bed volumes (BV) have been processed. The final effluent from the exchange unit is then held in the observation pond for further investigation or observation.

Resin Regeneration

After each 190 BV processed in the ion-exchange unit, the resin is regenerated using a conventional sodium chloride cycle consisting of

1) backwashing to remove accumulated particulate matter, 2) regeneration with a brine solution, and 3) a rinse to remove excess chloride before processing the next 190 BV.

Resin Restoration

In addition to the loss of exchange capacity due to ion replacement there is also a loss of exchange capacity due to the gradual buildup of organic and physical foulants. Organic fouling is due to the ability of the resin to remove negatively charged organic molecules and the inability of the regeneration process to re-exchange these organics as effectively as the inorganic anions. This removal characteristic accounts for the fractional ABS and COD reductions which are observed to take place in the exchange column. Laboratory studies have shown that these organic foulants can be almost completely removed by treating the resin with an organic solvent. Methanol has been used as the organic solvent in the present investigation because of its efficiency and economic availability.

Physical fouling refers to the accumulation of particulate matter which was not previously removed by filtration. This accumulation occurs at a very slow rate and this rate is directly related to the effectiveness of the resin backwashing process. Complete removal of the particulate matter during each regeneration cycle is not economically feasible, due to the excessive amounts of backwash water and time which would be needed. A hydrochloric acid bath (approximately 1.0 normal) has been used to effectively remove the particulate matter which is allowed to accumulate until resin efficiency is affected.

During the period covered by the August 1966 report (1), acid and methanol were used simultaneous in a single solution. The combined operation was called a resin restoration. During the last year resin restorations were made using acid and methanol separately, depending on the degree of physical and organic fouling. Significant economic gains were thus made and are discussed later in this report.

Sampling Program

To provide maximum control of all units in the treatment process and to study their operation it was necessary to plan and conduct a

comprehensive sampling program. The adopted sampling program schedule is shown in Table 1. As noted, in order to study the efficiency of each unit, influent and effluent samples were collected. Sampling intervals were selected based on the characteristics of each unit. During the conduct of resin efficiency studies it should be noted that the sampling frequency for the ion-exchange column was increased (approximately every 30 BV).

PILOT PLANT PERFORMANCE

Information and data obtained from the operation of the demonstration plant during the period from August 16, 1966 to September 15, 1967 is presented in this section. Material to be discussed includes 1) ion-exchange performance, 2) resin restoration techniques, 3) the removal of the various forms of phosphorus, 4) filtration studies, 5) virus removal studies, 6) mathematical studies, and 7) economic analysis.

Ion Exchange Performance

The initial phases of the demonstration project were devoted to the prediction of estimated pilot plant performance based on laboratory studies. The Demonstration Plant was constructed and operated with the objective of approaching the predicted performance on a pilot plant scale. The performance of the ion-exchange unit during past report periods has been compared to the predicted performance and discussed in previous progress reports (1) (2).

As previously indicated, one of the program objectives for the period covered by this report was to improve the demonstration plant performance. Towards that end, it was found that the performance of the ion-exchange column could be improved significantly by properly controlling of the activated sludge process and the sand filter which precede it.

Predicted Performance. As previously noted, prior to the operation of the demonstration plant, laboratory studies were conducted on the various treatment processes which were to be investigated on the pilot plant scale (3). These laboratory results were then used to estimate pilot plant performance. Data obtained from the laboratory studies for the ion exchange process have been summarized and are presented in Table 2.

TABLE 1

PILOT PLANT SAMPLING PROGRAM

Test [a]	Pilot Plant Components		
	Activated Sludge	Sand Filter	Ion Exchange
BOD	X		
COD	X	X	X
Suspended Solids	X	X	X
Phosphorus			
Ortho	X	X	X
Poly	X	X	X
Organic	X	X	X
Nitrate Nitrogen	X	X	X
ABS	X	X	X
Dissolved Oxygen	X		

[a] All tests were performed on the influent and effluent from each unit.

Actual Performance. The improved ion-exchange performance which was realized due to improved process control is clearly illustrated in Figures 2 and 3. Figures 2a and 2b show typical phosphorus removal efficiencies, both as percent and amount remaining in effluent, during two different periods of operation, while Figures 3a and 3b show typical nitrate nitrogen removals for the same time periods. The initial period of operation is represented by plant performance during the month of October 1966 and the final period of operation is represented by plant performance during the month of May 1967.

The most significant thing shown by these curves is the rapid loss of removal efficiency which occurred during the earlier periods of operation. During the October period the rapid loss of removal efficiency began at approximately 120 BV and 170 BV for phosphorus and nitrates. respectively. During the later periods of operation no significant loss of efficiency occurred for phosphorus or nitrate removal at 190 BV, or the end of the run. It is reasonable to assume that the length of run could be extended to 210 to 220 BV, which would reduce operating cost considerably.

The average efficiencies for 3000 BV processed beginning 7000 BV after the last restoration are shown for the initial and final periods of operation in Table 3. A comparison of Tables 2 and 3 indicate that the efficiencies obtained during the final periods of operation approach those predicted by laboratory results. As an example, the predicted phosphorus removal for approximately 8000 BV processed was 90 percent and the actual performance for approximately 8500 BV was 88.2 percent. At the same time the predicted nitrate removal was 85 percent while the actual removal was 82.3 percent.

Resin Restoration Techniques

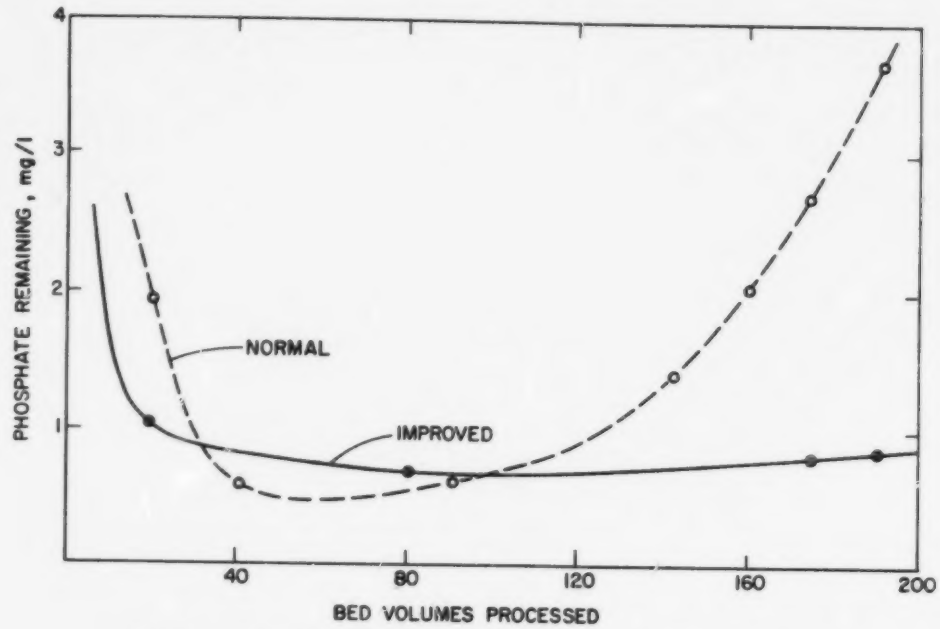
Based on the results presented in the last progress report (1), it was found that the ion-exchange resin became physically fouled before it became organically fouled. It was decided to attempt to separate the acid-methanol treatment into two restoration processes. The acid

TABLE 2

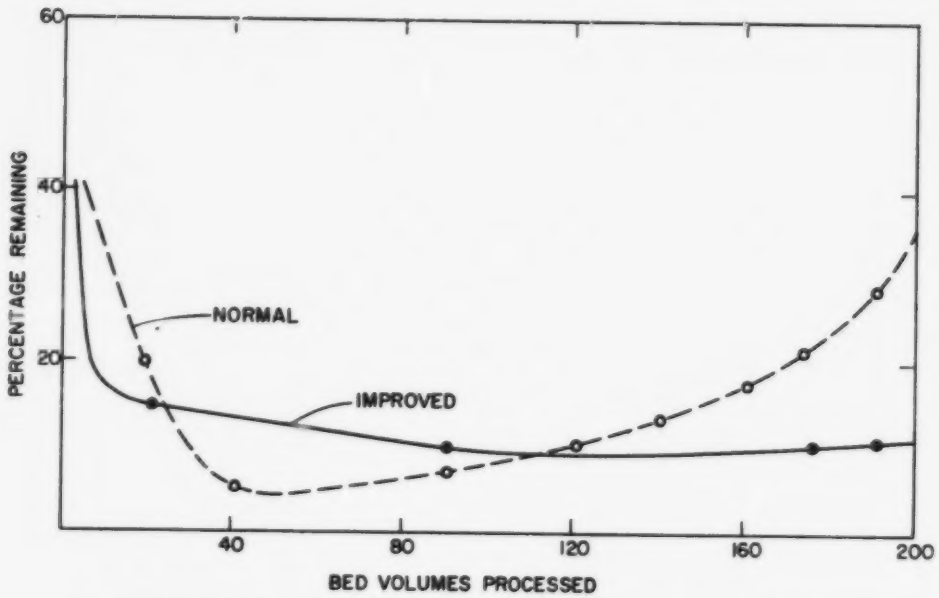
PREDICTED ION EXCHANGE PERFORMANCE [a]

Item	For 6000 BV		For 10000 BV	
	Average Effluent Conc., mg/l	Average Removal %	Average Effluent Conc.,mg/l	Average Removal %
Phosphorus	0.4	95	1.2	85
NO_3	2.0	90	4.0	80
ABS	0.25	80	0.5	60
COD	20	50	30	25

[a] Based on laboratory experimental studies reported in reference (3).

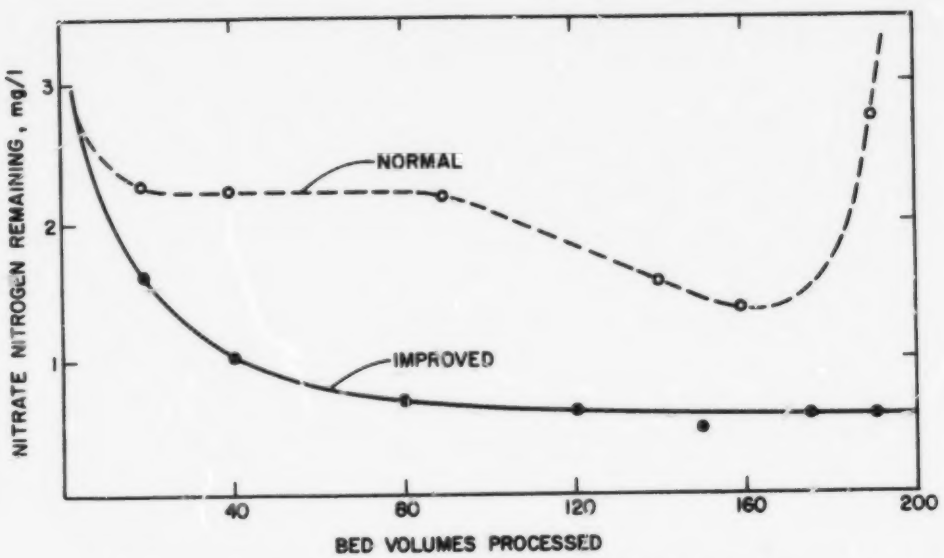


a) AMOUNT REMAINING VERSUS BED VOLUMES PROCESSED

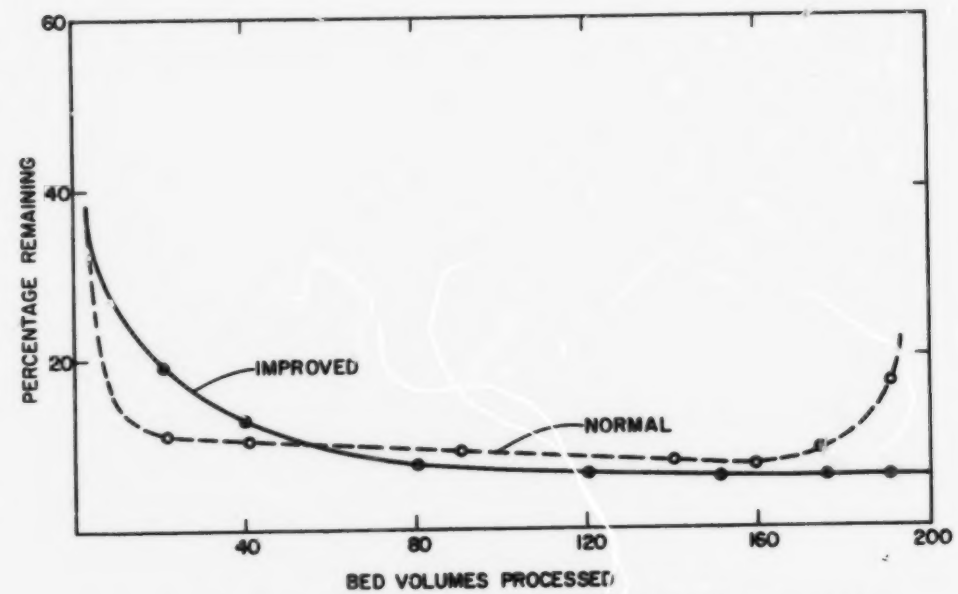


b) PERCENTAGE REMAINING VERSUS BED VOLUMES PROCESSED

FIGURE 2
ION EXCHANGE PERFORMANCE
BASED ON PHOSPHATE REMOVAL



a) AMOUNT REMAINING VERSUS BED VOLUMES PROCESSED



b) PERCENTAGE REMAINING VERSUS BED VOLUMES PROCESSED

FIGURE 3
ION EXCHANGE PERFORMANCE
BASED ON NITRATE NITROGEN REMOVAL

TABLE 3

ACTUAL ION EXCHANGE PERFORMANCE [a]

	Initial Period of Operation	Final Period of Operation	Percent Improvement
Average P in effluent	1.29 mg/l	0.59 mg/l	54.3
Average P reduction	74.7%	88.2%	18.1
Average NO_3 in effluent	1.73 mg/l	1.12 mg/l	35.2
Average NO_3 reduction	70.1%	82.3%	17.4

[a] Results are average efficiencies for 3000 bed volumes processed beginning 7000 bed volumes after the last restoration.

restoration would be used to restore the resin when it became physically fouled, and the methanol restoration would be used less frequently to restore organically fouled resin. It was previously estimated that the resin would have to have an acid-methanol restoration after no more than 14,000 BV had been processed. With more frequent use of the less expensive acid restorations approximately 20,000 BV can be processed between the more expensive methanol restorations. The savings which can be realized are significant and will be discussed later in this report. Table 4 shows the results of an acid restoration after 12,250 BV had been processed.

Based on the results of the current years pilot plant operation the following optimized restoration cycle has been suggested: a 1.0 normal HCL acid restoration every 10,000 BV and a methanol restoration every 20,000 BV. The methanol restoration and every other acid restoration could be done simultaneously to decrease the down time of the ion-exchange column.

Removal of Various Phosphorus Forms

There are thre major forms of phosphorus which are of importance. These are the orthophosphates, polyphosphates, and organic phosphorus. Of these three groups orthophosphate is the most readily available form for aquatic plant growth The polyphosphates must first be converted to orthophosphates to become available for plant growth and this requires time. Under normal receiving water conditions the time required for this conversion varies from a few hours to days (4). The major factors which effect the rate of conversion are temperature, pH, and the presence of bacterial or algal enzymes. Since organic phosphorus is generally very stable, it need only be considered when evaluating the potential long term effects of plant nutrients contained in reclaimed sewage effluents.

A more meaningful indication of nutrient problems associated with the phosphorus in reclaimed sewage effluent can be obtained by evaluating the separate forms of phosphorus. Although the organic form is not immediately suitable for algal metabolism it is nevertheless available for possible conversion to a suitable form by the action of bacteria and algae in the

TABLE 4

ACID RESTORATION OF ION EXCHANGE RESIN [a]

Item Removed	Removal Efficiency, %		Percentage Improvement
	Before Restoration	After Restoration	
PO_4	83.4	91.7	10.0
NO_3	76.5	83.6	9.3
ABS	47.9	59.0	23.2
COD	14.0	18.8	34.3

[a] 12,250 BV had been processed since last restoration.

receiving waters. Therefore, if treated wastewater is to be impounded, all of the previously mentioned forms of phosphorus may contribute to the growth of algae and other aquatic plants. On the other hand, if effluents containing phosphorus compounds are discharged into a rapidly moving stream the nutritive effects of these compounds would occur at various downstream locations. For example, the effects of the orthophosphates would be felt in the immediate vicinity of the point of discharge, while effects of the polyphosphates would not be felt for many miles downstream. The effects of the organic phosphorus would only be felt in the far reaches of the stream. Thus by evaluating the forms of phosphorus in an effluent it may be possible to establish a time and distance relationship between the amounts of phosphates present in each form and the possible plant nutrient problems which could occur. Table 5 shows typical ion-exchange removal efficiencies for the various phosphorus forms. An examination of Table 5 shows that while 0.60 mg/l of total phosphorus remains in the effluent only about half of this amount or 0.34 mg/l (ortho plus polyphosphates) need be considered when evaluating the immediate effects of the particular discharge on a stream. Also the percentage removal of ortho and polyphosphates averaged 91.5 percent when the total phosphorus removal was 88.2 percent.

Filtration Studies

Recognizing the importance of the filtration process as a means of improving the performance of the ion-exchange unit by removing suspended solids, studies were undertaken to establish the interrelationships between the process variables which control the effectiveness of the filtration process. The most important variables along with possible particle removal mechanisms have been summarized and are presented in Table 6. In order to study the interrelationships which exists between the process variables shown in Table 6, specialized pilot scale experimental equipment was designed and constructed, the details of which are presented in the following sections.

TABLE 5

ION EXCHANGE REMOVAL EFFICIENCY FOR VARIOUS PHOSPHORUS FORMS

Phosphorus Form	Amount in Effluent	Percent Reduction
Total	0.60 mg/l	88.2
Ortho	0.28 mg/l	92.0
Poly	0.06 mg/l	91.0
Organic	0.26 mg/l	73.6

TABLE 6

PARTICLE REMOVAL MECHANISMS AND FILTRATION PROCESS VARIABLES [a]

Removal Mechanisms	Process Variables
1. Straining	1. Filter media density
a. Mechanical	2. Filter media grain size
b. Chance contact	3. Filter media porosity
2. Sedimentation	4. Filter bed depth
3. Chemical Adsorption	5. Filtration depth
a. Bonding	6. Filtration rate
b. Chemical interaction	7. Available head
4. Physical Adsorption	8. Influent suspended solids concentration
a. Electrokenetic forces	9. Floc. strength and characteristics
b. Van der Waals forces	10. Fluid temperature, density, and viscosity
5. Adhesion	
6. Biological Activity	
7. Flocculation	

[a] These are all related to characteristics listed under filter operation and design.

Experimental Equipment. The experimental equipment and apparatus used for the pilot scale filtration studies is shown schematically in Figure 4. The equipment is comprised of 1) settled effluent pumps, 2) constant head tank, 3) experimental filter columns and appurtenances, 4) effluent pumps and 5) monometer board (not shown).

The experimental process flow sheet may be described as follows. Settled sewage effluent is pumped from the effluent trough of the secondary activated sludge treatment unit to the constant head tank. From this unit, which is used to control the head on the filters, the effluent is applied to the three experimental filters which are operated in parallel. Effluent from the filter columns is discharge to the plant sewer. It should be noted that the filter effluent pumps, which are of the positive displacement flexible impeller type, also serve to control the flowrate through the filter columns. Minor variations in the flow rate can be controlled by adjusting the valve located on the discharge side of the pump.

In order to conduct a wide variety of filtration experiments, the columns were constructed so that the test sections (see Figure 4) would be removable. This allowed test sections to be interchanged as required. For example, in one series of tests, the filtration process was continued until the medium in the filter columns became completely clogged, whereupon the filter test section was removed and the granular material within the test section was layered and the amount of suspended material retained in the various layers was determined.

To study concentration variations at various depths within the filter, sample collection taps were located at various intervals throughout the test section (see Figure 4). Test samples were withdrawn through hypodermic needles inserted into the column at the tap locations. Dual manometer taps were also located at the same level as the sampling taps.

Experimental Studies. Although a number of filtration experiments have been completed, the results of all of these experiments have not been, to date, fully analyzed. Therefore, this presentation will be limited to a discussion of the first series of experiments.

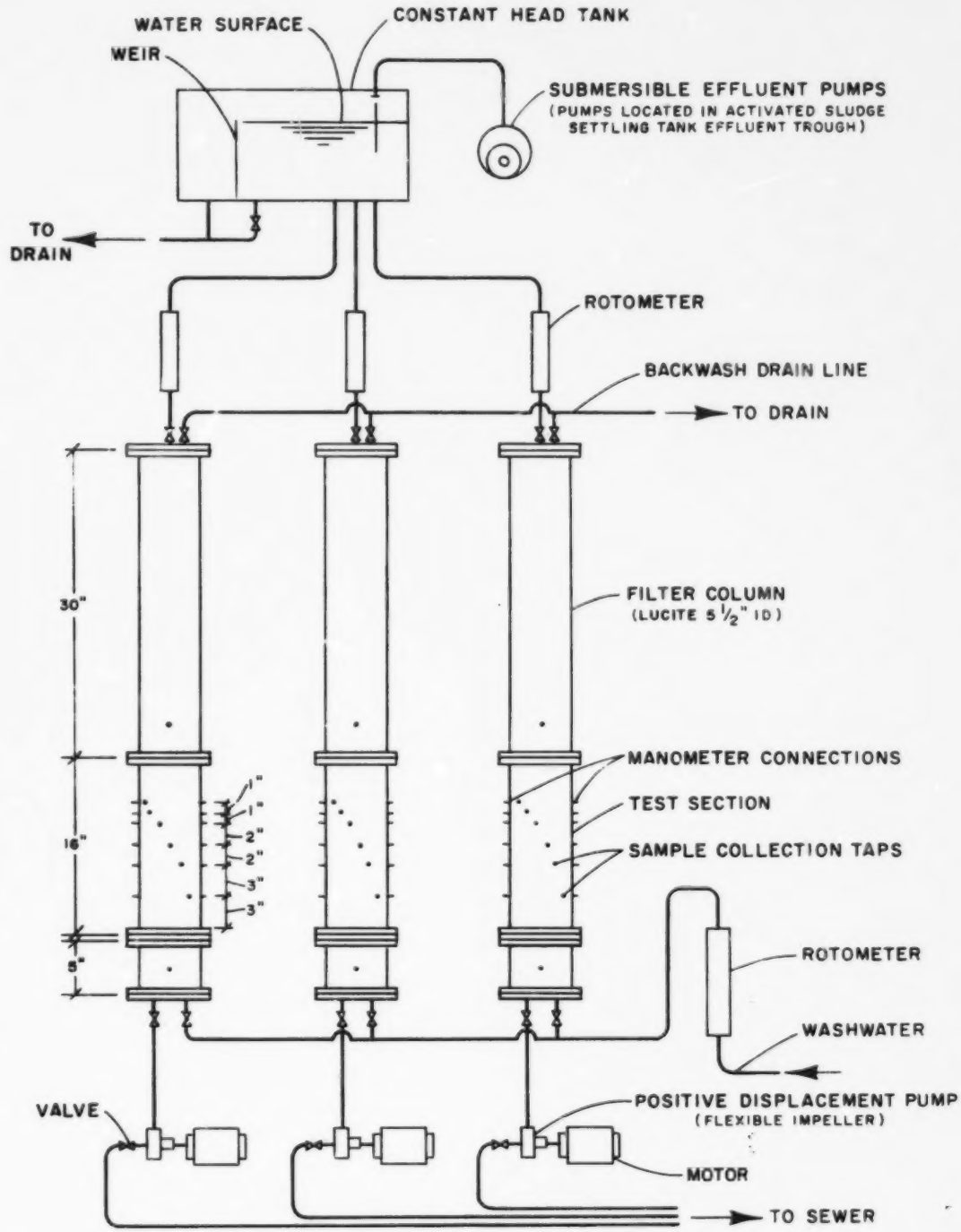


FIGURE 4
FILTRATION EQUIPMENT USED
FOR EXPERIMENTAL STUDIES

The objective of this series of experiments was to study the effects of the filtration process variables (see Table 6) on removal efficiency when filtering untreated settled activated sludge effluent, and to gather preliminary data on the removal mechanisms responsible for the action of the filter. As it was possible to control most of the process variables listed in Table 6, the effect of filtration rate and sand size were the first variables selected for detailed study. To determine meaningful relationships between these variables, filtration rates of 2.0, 5.8, and 10 gpm/sq. ft., and sand sizes of 0.488, 0.681, 0.978, and 1.275 mm were selected for the experimental studies.

Results of Experimental Studies. For ease of presentation the results of the experimental studies have been summarized and are presented graphically in Figure 5 through 9.

Typical results of experiments designed to demonstrate the effect of filtration rate on removal efficiency are shown in Figure 5. For a given sand size (0.681 mm), varying the filtration rate had relatively little effect on the initial suspended solids removal characteristics of the filter bed. This was observed for each of the sand sizes studied.

Results of experiments designed to illustrate the effect of sand size on filtration removal efficiency are shown in Figure 6. For a fixed flow rate (5.8 gpm/sq. ft.), varying the size of the filtering material had a significant effect on the initial suspended solids removal efficiency of the filter bed. This effect was also noted for each of the filtration rates studies.

Rather than present all of the data obtained during this study in repetative figures, such as Figure 5 and 6, the data has been cross correlated and is graphically summarized in Figure 7. Here, the relations between both sand size and filtration rate to removal efficiency are shown. The effect of each variable can be noted by observing the slope of the removal efficiency lines. For example, if the removal efficiency lines were vertical, the rate of filtration would have no effect on the removal efficiency.

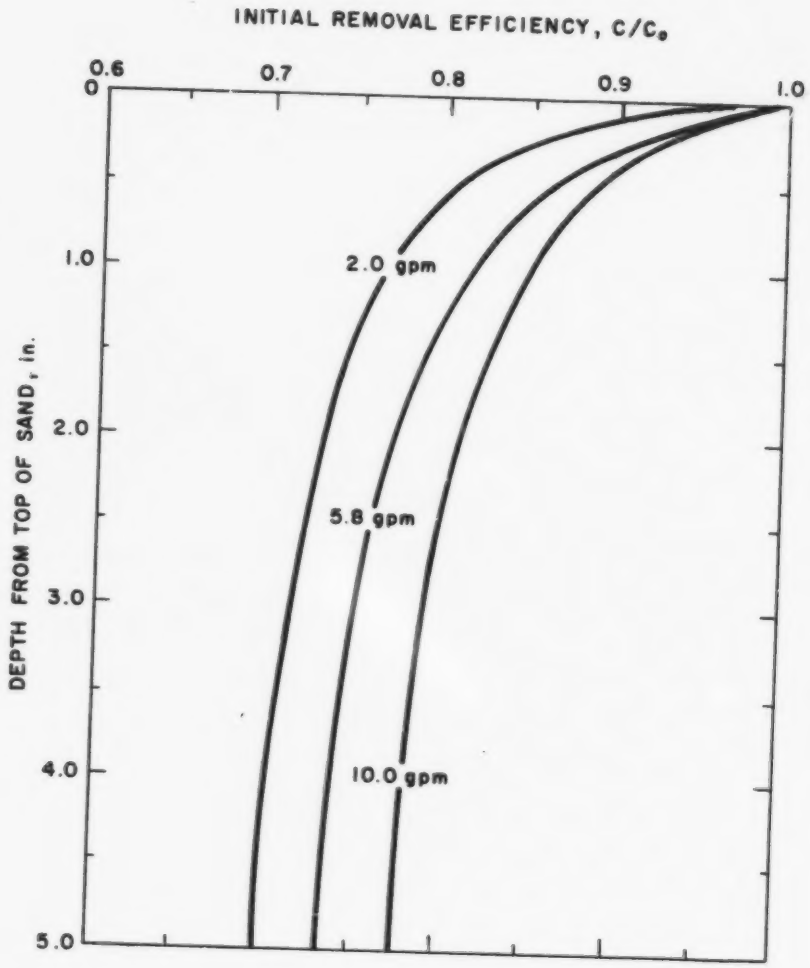


FIGURE 5
EFFECT OF FILTRATION RATE ON REMOVAL EFFICIENCY
(Sand Size 0.681 mm.)

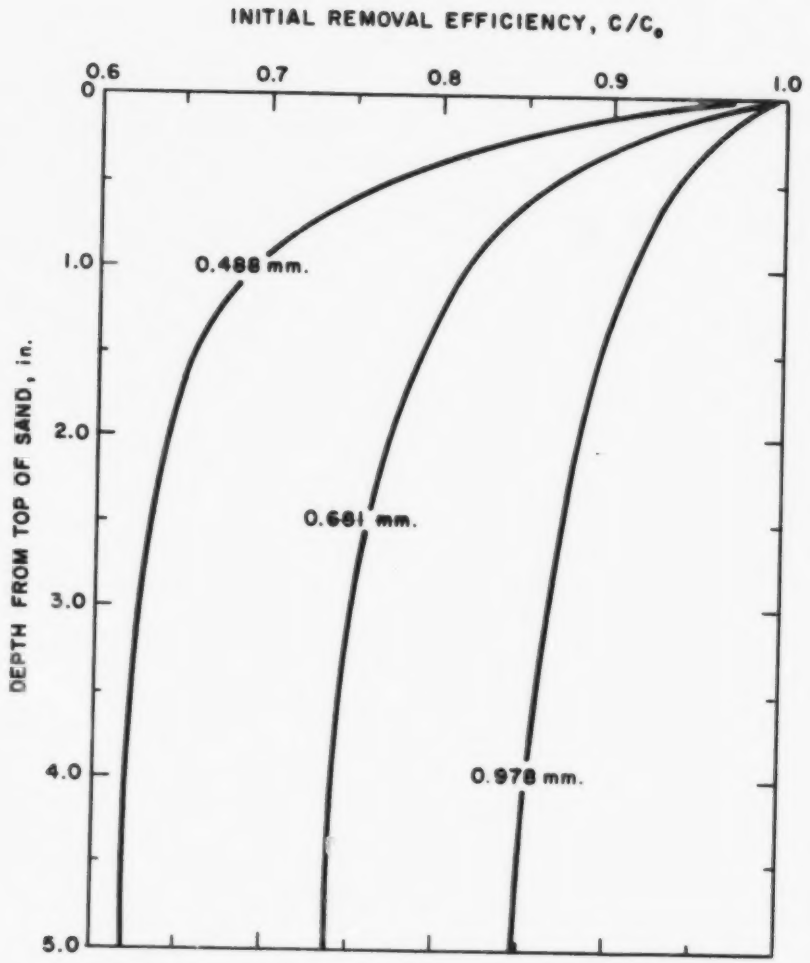


FIGURE 6
EFFECT OF SAND SIZE ON FILTRATION REMOVAL EFFICIENCY
(Filtration Rate 5.2 gpm/sq. ft.)

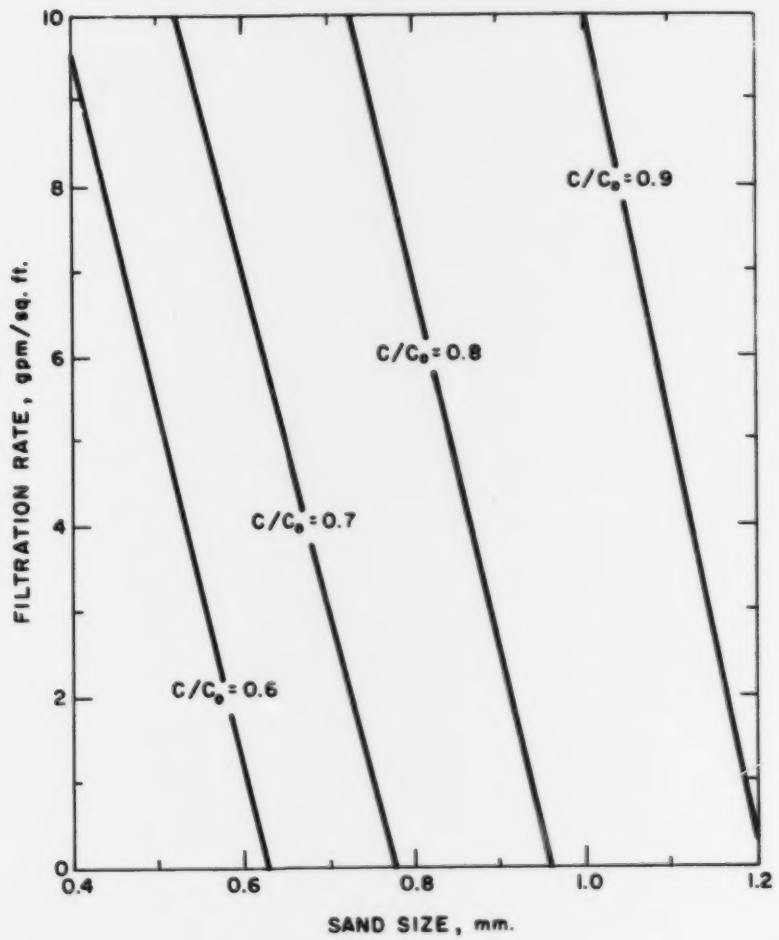


FIGURE 7
EFFECT OF FLOW RATE AND SAND SIZE
ON FILTRATION REMOVAL EFFICIENCY

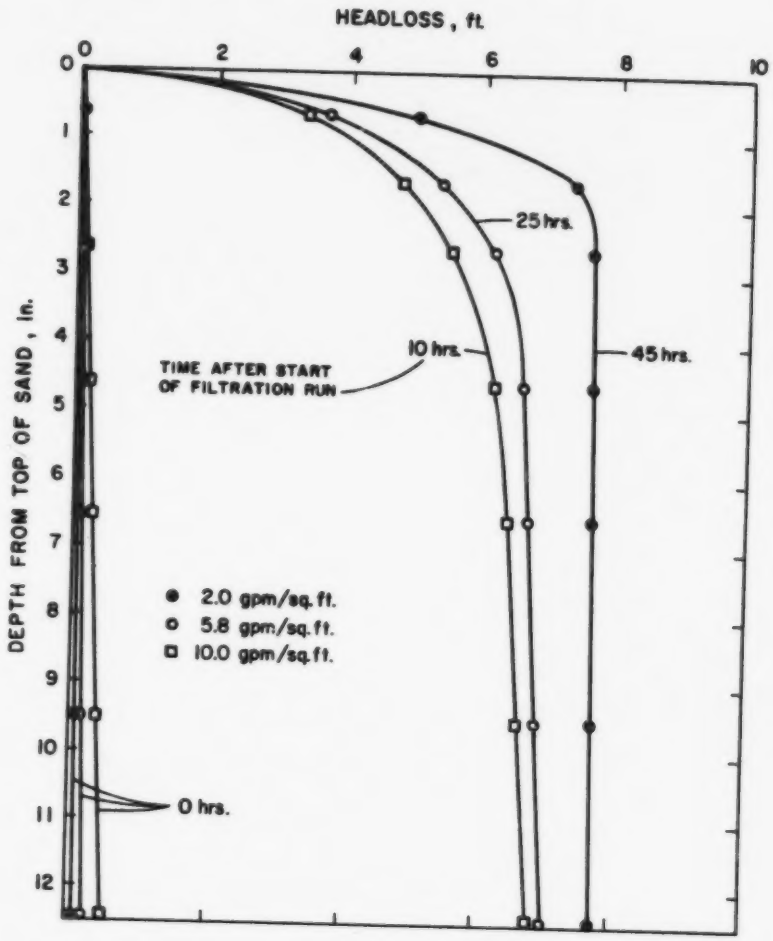


FIGURE 8
HEADLOSS VERSUS DEPTH FROM TOP
OF SAND FOR VARIOUS FILTRATION RATES
(SAND SIZE 1.27 mm)

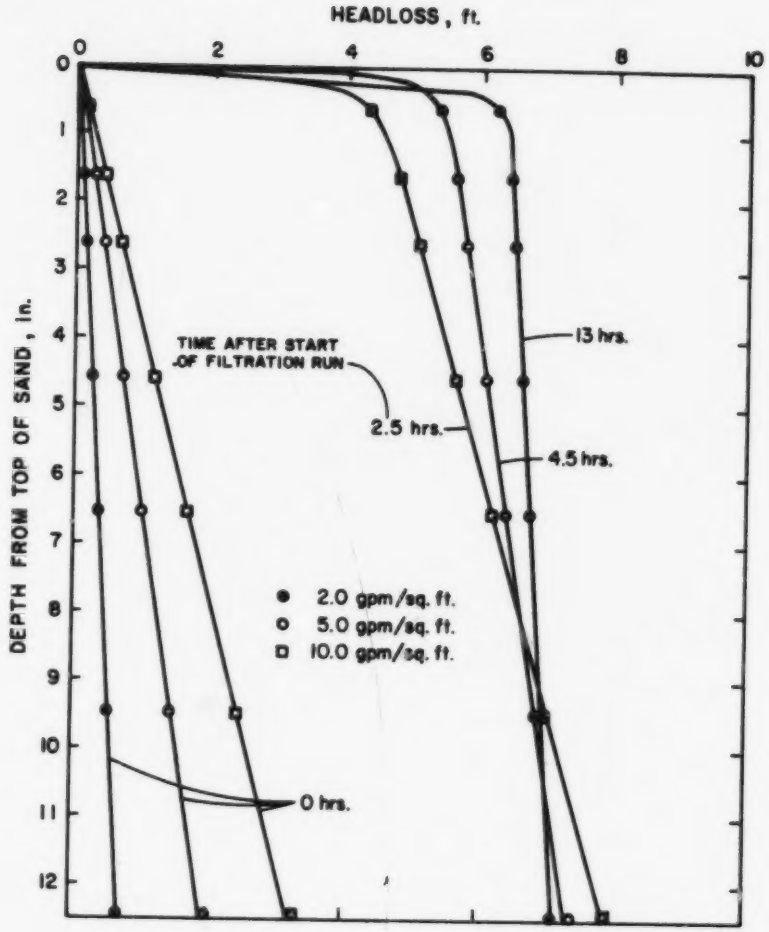


FIGURE 9
HEADLOSS VERSUS DEPTH FROM TOP
OF SAND FOR VARIOUS FILTRATION RATES
(SAND SIZE 0.488 mm)

The effect of the rate of filtration and sand size on headloss is shown in Figures 8 and 9. Headloss as a function of depth in the column for various filtration rates and a constant sand size of 1.27 mm is shown in Figure 8. Headloss due to the passage of clean tap water through the columns is depicted by the three curves on the left hand side of this figure. The other three curves shown represent the situation after the column has been in operation for the time indicated on the curves.

By noting the shape of the curves it is possible to obtain a qualitative indication of the manner in which the material removed from the supernatant is distributed throughtout the sand column. For example, for the slow filtration rate (2.0 gpm/sq. ft.) it can be seen that almost all of the removal is occurring in the upper 2-1/2 inches by comparing the slope of the curve below this point to the slope of the freshwater filtration curve. However, for the other filtration rates it can be seen that significant removal of material is taking place to a depth of about 6-1/2 inches and 4-1/2 inches for filtration rates of 10 and 5.0 gpm/sq. ft., respectively.

By comparison, when the sand size is decreased to about 0.488 mm it can be seen from Figure 9 that regardless of the rate of filtration most of the removal is confined to the upper 1/2 to 1 inch of the filter column. The implication of these findings is currently under investigation.

Virus Removal Studies

The removal of viruses is becoming of more concern as the application of treated wastewater for groundwater recharge and reuse continues to increase. In this connection, the objectives of the virus removal studies to be discussed in this report were to determine the effectiveness of the filtration process in removing viruses and to study the effects of coagulants and coagulant aids on virus removal efficiency using sand filters. To meet the stated objectives both static (batch) and dynamic (column) studies were conducted. Batch studies were used to determine virus-sand adsorption characteristics with and without coagulants and coagulant aids. Column studies were used to study the dynamic removal of virus on a pilot plant scale which would simulate actual operating conditions for a full scale plant.

Procedures for conducting both static and dynamic tests have been documented and may be found in reference (7). Details on the materials, procedures, and equipment used in these studies are presented below.

Materials and Procedures. Virus bacteriophage T-2 and phage 65 were used in the current study because of their ability to form easily counted plaques with Escherichia coli (1). Using the T-2, phage 65, and Escherichia coli the incubation time was reduced from the normal 18 to 24 hours to 6 hours at 37° C. The initial cultures of the T-2 and phage 65 viruses and all of the Escherichia coli strains were obtained from the Medical Microbiology Department of the School of Medicine, Stanford University. The cultures were then maintained in the microbiology laboratory of the Water Quality Control Research Laboratory at a temperature of 2° C.

Coagulants and Coagulant Aids. In earlier studies (5)(6) Alum ($Al_2(SO_4)_3 \cdot 18H_2O$) and ferric chloride ($FeCl_3$) were used separately and together to investigate their effects on virus adsorption on sand. Also several polyelectrolytes were used as coagulant aids to determine their effect on virus adsorption on sands. The positively charged or cationic polymers used were Separan C-90 and C-120. The negatively charged or anionic polymer used was AP-30 and the neutral (no charge) polymer used was NP-10. Based on the results of the earlier studies, it was found that a combination of C-120 polyelectrolyte and alum gave the best virus removal efficiencies. The current studies which focused on the effects of coagulants and coagulant aids were conducted using combinations of C-120 polyelectrolyte and alum.

Results of Batch Studies. Using 200 ml of water, 100 grams of Delmonte sand (effective size = 0.49 mm and uniformly coefficienct = 1.29), and various dosages of Alum and C-120 polyelectrolyte, batch studies were conducted to determine the effect of chemical additions on virus adsorption. The results of batch studies conducted without the addition of either alum or polyelectrolyte indicated that the virus adsorption efficiency of the sand used in these experiments was equal to about 40 percent ($C/C_o = 0.6$). As shown in Figure 10, using various alum dosages without the addition of polyelectrolyte, the average virus adsorption efficiency was increased to

about 70 percent ($C/C_o = 0.3$). The addition of various dosages of polyelectrolyte had little further effect on the virus adsorption efficiency, as indicated by the results shown in Figure 10. A slight increase in the removal efficiency was observed at higher polyelectrolyte dosages (about 0.3 ppm) but these combinations of alum and polyelectrolyte were not investigated further because of economic considerations associated with the high cost of polyelectrolytes. In practice, an economical analysis would be required to determine if the incremental improvement in removal efficiency would be justified by the increased cost.

Results of Column Studies. Using small lucite columns, laboratory studies were conducted to determine the effect of filtration rate on the removal of viruses through sand filters. The filter sand used in these studies consisted of a mixture 95 percent 9.16 mm and 5 percent 0.11 mm diameter Monterey white silica sand. The results of these studies have been summarized and are presented graphically in Figure 11. As shown in Figure 11 the time required to reach saturation is a function of the filtration rate. For filtration rates up to about 5.2 gpm/ft^2 the time required to reach saturation may be formulated as follows:

$$t\ (\text{min}) = 247 - 145 \ln (\text{filtration rate, gpm/ft}^2) \qquad (1)$$

The constants in this equation appear to depend on sand size, and initial virus concentrations.

Another series of filtration runs performed during the period covered by this report were designed to investigate the effect of the addition of coagulants and coagulant aids. The filter columns used for these studies, shown in Figure 4, have been previously described in the section dealing with filtration studies. The sand used in these filtration experiments had an effective size of 0.28 mm and a uniformly coefficient of 1.18. The depth of sand in the filter was 16 inches. The feed solution for these tests contained 50 ppm alum, 0.2 ppm cationic (C-120) polyelectrolyte, and a concentration of viruses equal to 7.33×10^6 per milliliter.

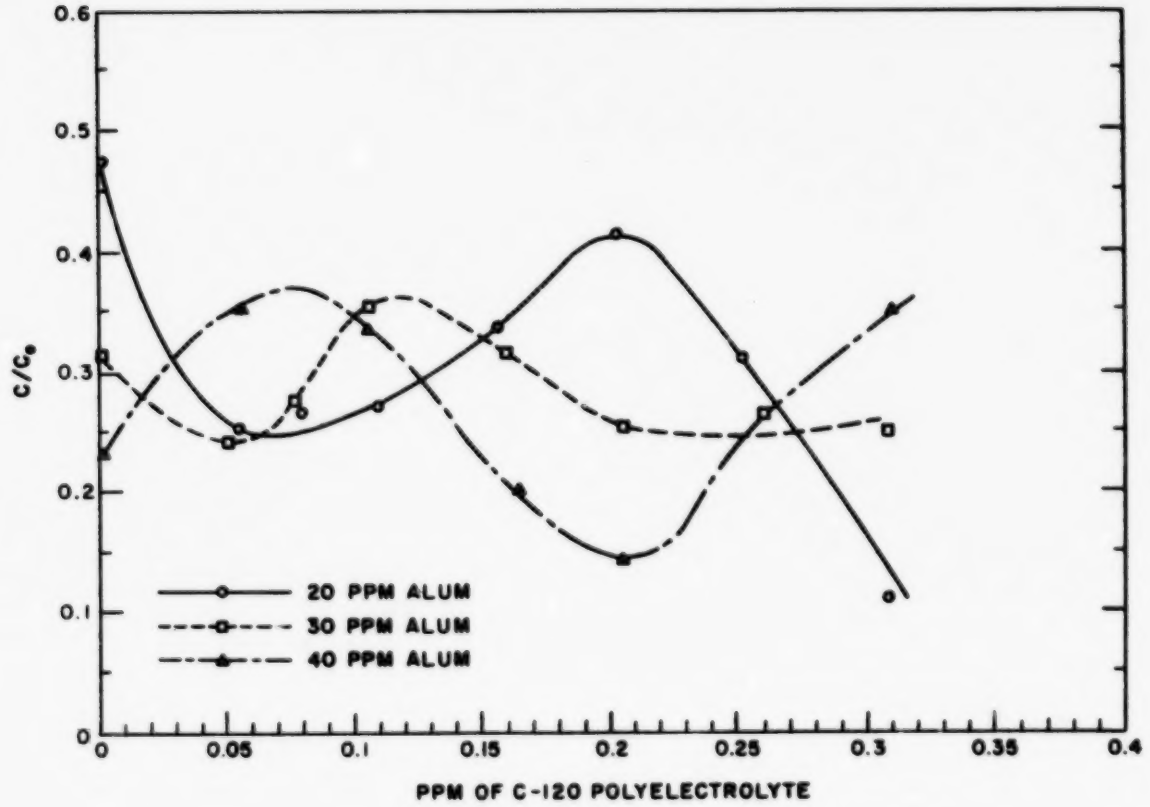


FIGURE 10
RESULTS OF BATCH VIRUS ADSORPTION STUDIES
USING ALUM AND C-120 POLYELECTROLYTE

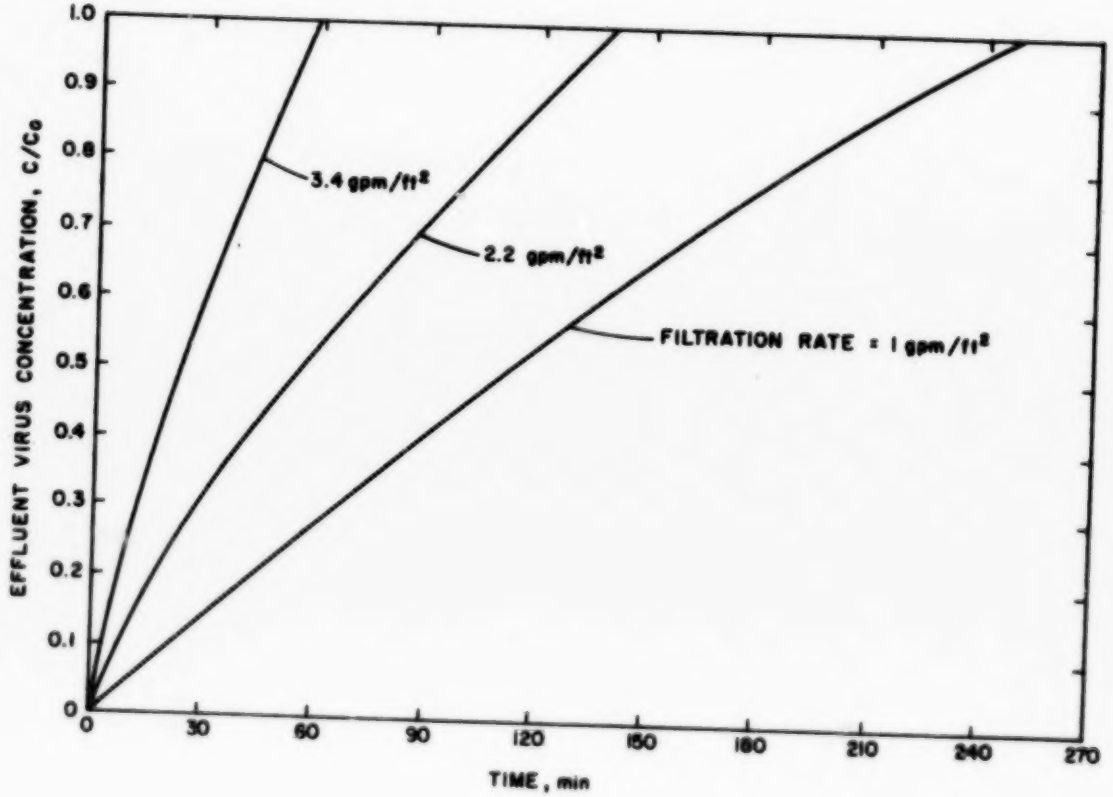


FIGURE 11
RATIO OF EFFLUENT CONCENTRATION (C) OF VIRUSES TO FEED CONCENTRATION (C_0) VERSUS TIME FOR VARIOUS FILTRATION RATES

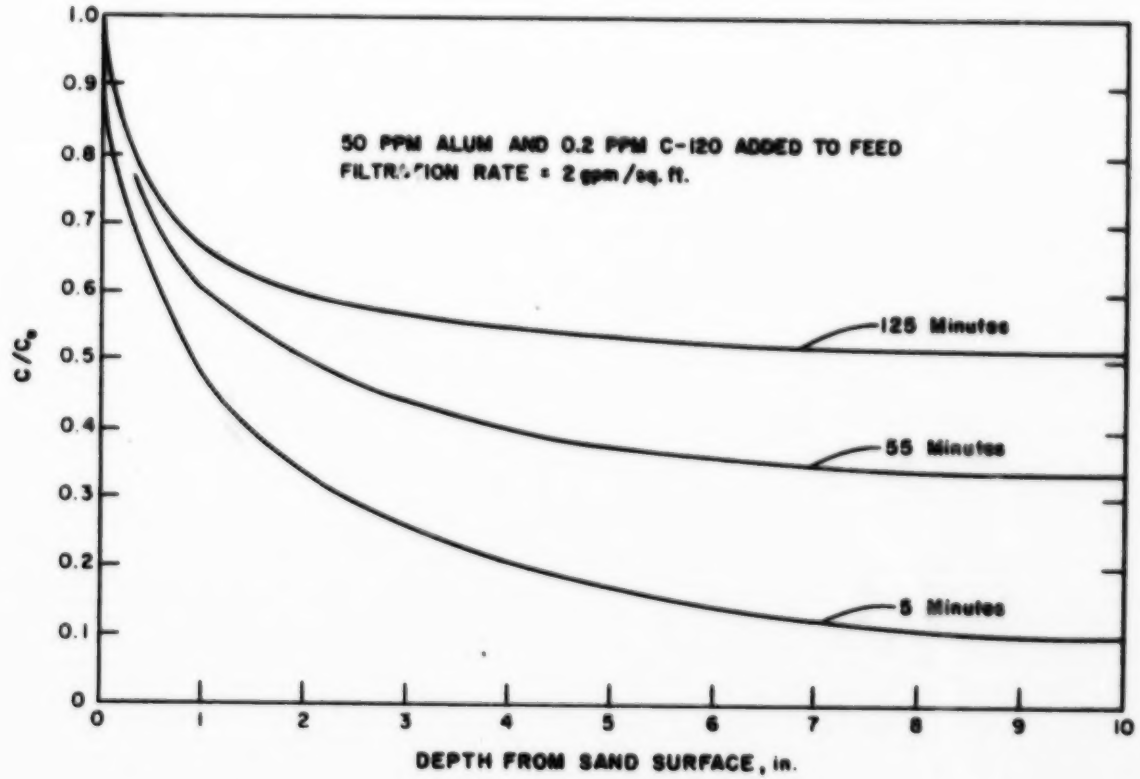


FIGURE 12
VIRUS REMOVAL AS A FUNCTION
OF FILTER DEPTH AND TIME

The solution was fed into the column at a rate of 2 gpm/ft^2 for 3 hours. The results of a typical test run are presented in Figure 12 which is a plot of the concentration of virus versus depth from sand surface at various times. As shown in Figure 12 the filter rapidly loses its ability to remove viruses. After only 5 minutes, the concentration of viruses coming through the filter was approximately equal to 10 percent of the initial feed concentration. After 2 hours the effluent contained over 50 percent of the viruses in the feed solution.

The conclusion drawn from these experiments is that conventional filtration is not a very effective process for the removal of viruses. Further, the addition of various dosages of alum and polyelectroyte does not significantly effect the virus removal efficiency of the filtration process.

Mathematical Studies

As previously mentioned, one of the main objectives of the current years study was to improve the performance of the demonstration plant by properly controlling the activated sludge and sand filtration processes which precede the ion-exchange unit. To provide optimum control it was decided that an attempt should be made to describe each process by a mathematical or graphical model. These mathematical or graphical "models" would then be incorporated into an overall model which could be used to control each unit in a way which would result in the most efficient overall plant performance. During the current period of operation, mathematical and statistical techniques were used to develop a model which could be used to evaluate the phosphorus removal efficiency of the ion-exchange column.

Evaluation of Phosphorus Removal by Ion Exchange. As presented in the section on Economic Analysis, a major portion of the cost of the renovation process is for the chemicals required for the regeneration and restoration of the ion-exchange resins. It was felt that a significant reduction in chemical costs could be realized if an effective method of coordinating the regeneration and restoration cycles could be developed. To this end, a mathematicl model has been developed which incorporates mean phosphorus

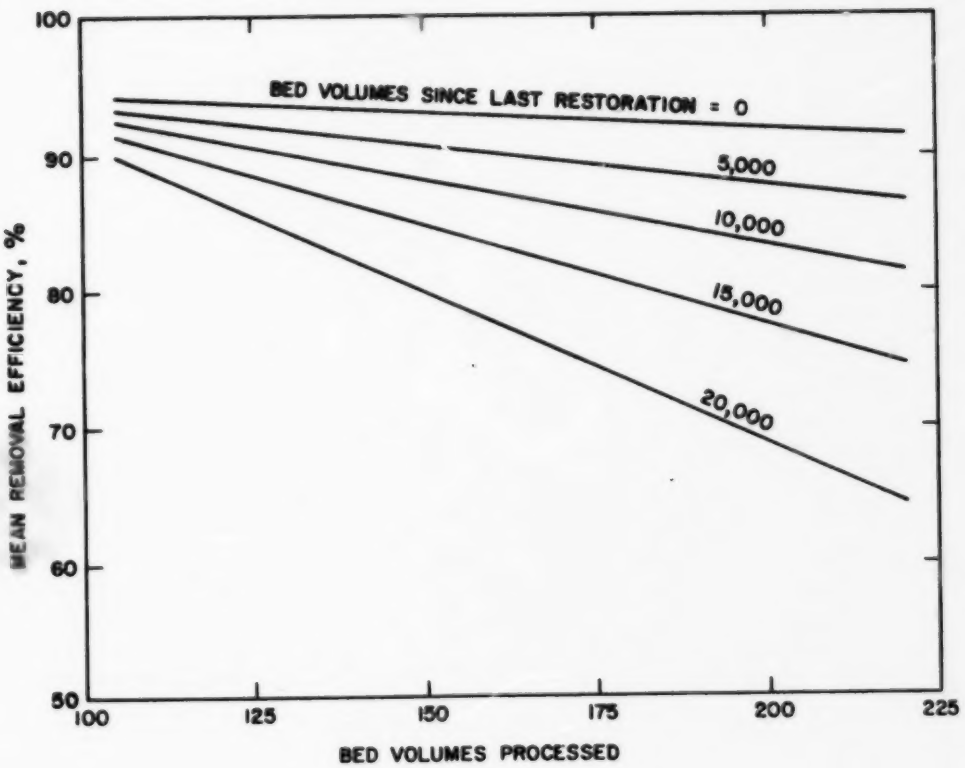


FIGURE 13
ION EXCHANGE PERFORMANCE AS A FUNCTION OF RESIN AGE
BASED ON PHOSPHORUS REMOVAL

removal efficiency, resin regeneration cycle, and resin restoration cycle. The model, presented in graphical form in Figure 13, indicates that it is possible to adjust the length of both the regeneration and restoration cycles to provide any required removal efficiency. If, for example, a minimum removal efficiency of 85 percent is required and 10,000 BV have been processed since the last restoration the number of BV which can be processed between regenerations would be about 180 BV.

Economic Analysis

In the previous progress report (1) a revised cost estimate was prepared which indicated the savings in chemical costs due to the removal of the diatomaceous earth filter, the abandonment of the bentonite ion-exchange restoration process, and the reduction in the volume of methanol needed for adequate restoration. Also there was a reduction in the estimated capitol recovery cost based primarily upon the equal or better resin performance achieved at higher flow rates and on the use of the sand filter instead of the diatomaceous earth filter.

Because of the improved ion-exchange process performance obtained during the past year, additional savings in chemical costs were estimated. These savings result from the increased number of bed volumes which can be processed between salt (NaCl) regeneration and the previously discussed optimized restoration cycle. Table 7 shows the original cost estimate, the cost estimate presented in the last report and the current cost estimate.

TABLE 7

COST ESTIMATES FOR NUTRIENT REMOVAL (PHOSPHORUS AND NITRATE)

Item	Cost,[a] cents/1000 gal. Original Estimate[b] 1965	Revised Estimate 1966	Revised Estimate 1967
Resin Replacement	2.75	2.75	2.75
Salt	3.58	3.58	2.68 to 2.86
Bentonite	0.10	-	-
Methanol	2.50	1.00 to 1.75	0.70 to 1.23
Hydrochloric Acid	0.13	0.19 to 0.25	0.27 to 0.35
Sodium Hydroxide	0.09	0.00 to 0.09	0.00 to 0.09
Diatomaceous Earth	0.52	-	-
Sub-total:	9.67	7.52 to 8.42	6.40 to 7.28
Labor and Overhead	6.60	6.60	6.60
Power	1.43	1.43	1.25
Capitol Recovery Costs[a]	3.05	1.55 to 2.30	1.55 to 2.30
Total:	20.75	17.10 to 18.75	15.00 to 17.43

[a] Includes sand filter, ion exchange unit, chemical storage, and plant site development.

[b] From reference (8).

CONCLUSIONS

Conclusions derived from the operation of the pilot plant and special investigations conducted during the past thirteen months include:

1. Phosphate and nitrate removal efficiencies approaching those predicted by laboratory experiments can be obtained by properly controlling each process of the wastewater renovation plant.

2. Sustained removal efficiencies of phosphates and nitrates averaging 89.2 and 82.3 percent, respectively, can be achieved.

3. Essentially complete restoration of physically and organically fouled ion-exchange resins can be accomplished with acid and methanol restoration treatments, respectively.

4. A more meaningful indication of possible nutrient problems associated with the phosphorus in reclaimed sewage effluent may be obtained by evaluating the separate forms of phosphorus.

5. The effect of the rate of filtration on filter efficiency is small compared to the effect of sand size, when filtering settled effluent from the activated sludge treatment plant.

6. For the sand sizes and filtration rates normally used for the filtration of settled secondary effluent, about 75 to 90 percent of the total amount of material removed by the filter is contained in the upper one inch of the sand bed.

7. A linearized mathematical or graphical model of the overall pilot plant process could be used to minimize operation costs by optimizing the operation of the individual components.

8. The total estimated cost of combined nutrient removal (phosphorus plus nitrogen) using an anion exchange process is about 15.00 to 17.43 cents per 1000 gallons. Of the total, chemical costs will vary between 6.40 to 7.28 cents per 1000 gallons.

FUTURE OBJECTIVES

Future demonstration project objectives are:

1. To continue to operate the demonstration plant, on a routine basis, to further define the characteristics, advantages, and limitations of filtration and ion-exchange techniques in the renovation of wastewater.

2. To determine the control and quality implications of turbidity information, relative to wastewater renovation.

3. To investigate in depth the interrelationships between the parameters and variables which govern the effectiveness of the filtration process as applied to the filtration of sewage effluent.

4. To investigate the applicability of the filtration process for the renovation of sewage effluents which have been chemically treated for the removal of phosphates and suspended solids.

5. To study the feasibility of vacuum filtration for processing the sludge obtained from the precipitation of phosphates in secondary effluent with lime and other chemicals.

6. To explore in depth the implications and potential benefits resulting from the filtration of sewage treatment effluents, both in the context of the ion-exchange renovation, and also an an independent means of effluent quality improvement.

7. To prepare a final project report summarizing the results obtained during the past five years from the conduct of specialized studies and the operation of the demonstration plant.

REPORTS AND PUBLICATIONS

During the past year, three papers dealing with various aspects of wastewater reclamation or renovation have been prepared. The first of these reports entitled "Ultimate Disposal of Contaminants from Wastewater Reclamation Processes" (9) was co-authored by Rolf Eliassen and George Bennett and was presented before the California Water Pollution Control Association meeting at Long Beach, California on May 4, 1967. The second report entitled "Renovation of Domestic and Industrial Wastewater" (10) was also co-authored by Rolf Eliassen and George Bennett and was submitted to the International Conference on Water for Peace which was held in Washington, D. C. on May 23, 1967. The title of the third paper was "Chemical Processing of Wastewater for Nutrient Removal" (11); it was co-authored by Rolf Eliassen and George Tchobanoglous and was presented at the 40th Annual Conference of the Water Pollution Control Federaion in New York City, New York on October 12, 1967.

The objective persued by the project staff in writing and presenting these papers and reports is to promote the interchange of ideas, data, and information on the general subject of wastewater renovation. Keeping this objective in mind, portions of these papers have been abstracted for presentation in this present report. The portions of these reports which have been abstracted deal with 1) reclamation process materials mass balances, 2) ultimate disposal methods and costs, and 3) a comparison of the alternative removal processes for nitrogen and phosphorus. This data and information will serve as a useful adjunct to the data and information presented in this present report and other reports dealing with wastewater reclamation. The abstracted material was taken from the first and third of the aforementioned reports.

Quantities of Contaminants--Abstracted from Reference (9).

Processing of the liquid phase of wastewaters for removal of suspended. colloidal and dissolved solids, both organic and inorganic, is one phase of advanced waste treatment. The other phase involves

disposal of the removed contaminants. This could be a formidable problem for large plants located in some areas of the country. Organic matter may be subjected to the digestion process, but inorganic substances must be handled in another manner which might be termed ultimate disposal. Before considering this subject it will be necessary to determine the quantities of contaminants, both liquid and solids, which may be expected from the various advanced waste treatment processes previously outlined. The following discussion considers the flow diagrams and mass balances for the wastes produced in plants having a unit capacity of 1 mgd.

Coagulation-Sedimentation for Phosphorus Removal. Phosphorus removals accomplished using the process flow sheet shown in Figure 14 depend upon the amount of coagulants used, together with coagulant aids. A lime dosage of at least 300 mg/l will normally be required; this will produce great quantities of waste sludge. The amount and characteristics of this sludge will vary among plants. An average solids content of the combined organic and lime sludges could approximate four percent. Using this value, and assuming a lime dosage of 300 mg/l, the calcium carbonate produced per mg will contain approximately 3330 pounds of sludge plus 170 pounds of removed particulates. This means that 10,500 gallons of four percent sludge per mg constituting an additional 0.35 pounds of solids per capita per day, would have to be handled. This is almost double the normal solids content of 0.20 pounds per capita per day in municipal wastewaters.

The coagulation-sedimentation sludge will have only about 3.5 percent volatile matter; therefore, conventional disposal of anaerobic treatment does not appear desirable. Efforts have been made to develop a workable and economical technique for re-calcining the waste lime sludge for reuse. This serves two useful purposes in that in minimizes the amount of product for ultimate disposal and reduces the cost of coagulant chemicals significantly. Liquid wastes would be eliminated. It has been estimated that the re-calcining operation will permit satisfactory operation with fresh lime makeup as low as 25 to 35 percent of

the total dose (i.e., 30 percent fresh lime, 70 percent re-calcined lime). A mass balance for the secondary and tertiary processes using coagulation-sedimentation with lime recovery, followed by filtration is presented in Figure 14.

Process wastes which will require ultimate disposal consist of air-borne wastes from calcine furnace stack emissions and dry re-calcined lime which would have to be wasted to maintain the desired balance between fresh lime and re-calcined lime. No information is available at present regarding the stack emissions. Current concern regarding pollution of the air suggests some investigation of this aspect of lime sludge reclamation, since the necessity of including air pollution control equipment along with the calcine furnace could significantly affect the overall process economics.

Anion Exchange for Nitrogen and Phosphorus Removal. The flow-sheet in Figure 15 contains estimates of the waste brines which will be subjected to ultimate disposal. Filtration is indicated as an intermediate unit operation between the activated sludge plant and the ion-exchange column. Other wastes will occur from resin regeneration and from the periodic cleaning of the resin. These wastes will contain mainly dissolved salts. Chemical treatment for the restoration of the resin involves wasting small quantitites of methanol, caustic, and acid at infrequent intervals, with no discharge exceeding 200 gallons of chemical per million gallons of effluent processed. These could be discharged to the primary portion of the treatment plant.

Most of the volume of waste produced comes from the regeneration of the anion exchange resin with sodium chloride. The resin is exposed to a brine solution (11 percent salt) which has been used originally in the prior regeneration cycle. This brine is then discarded (thereby becoming a waste product) and the column is put back on line. The first tne percent of this rinse has a chloride content averaging 10,000 mg/l. The balance of the rinse water will have an average chloride concentration of about 450 mg/l. In an operating system each mg treated would

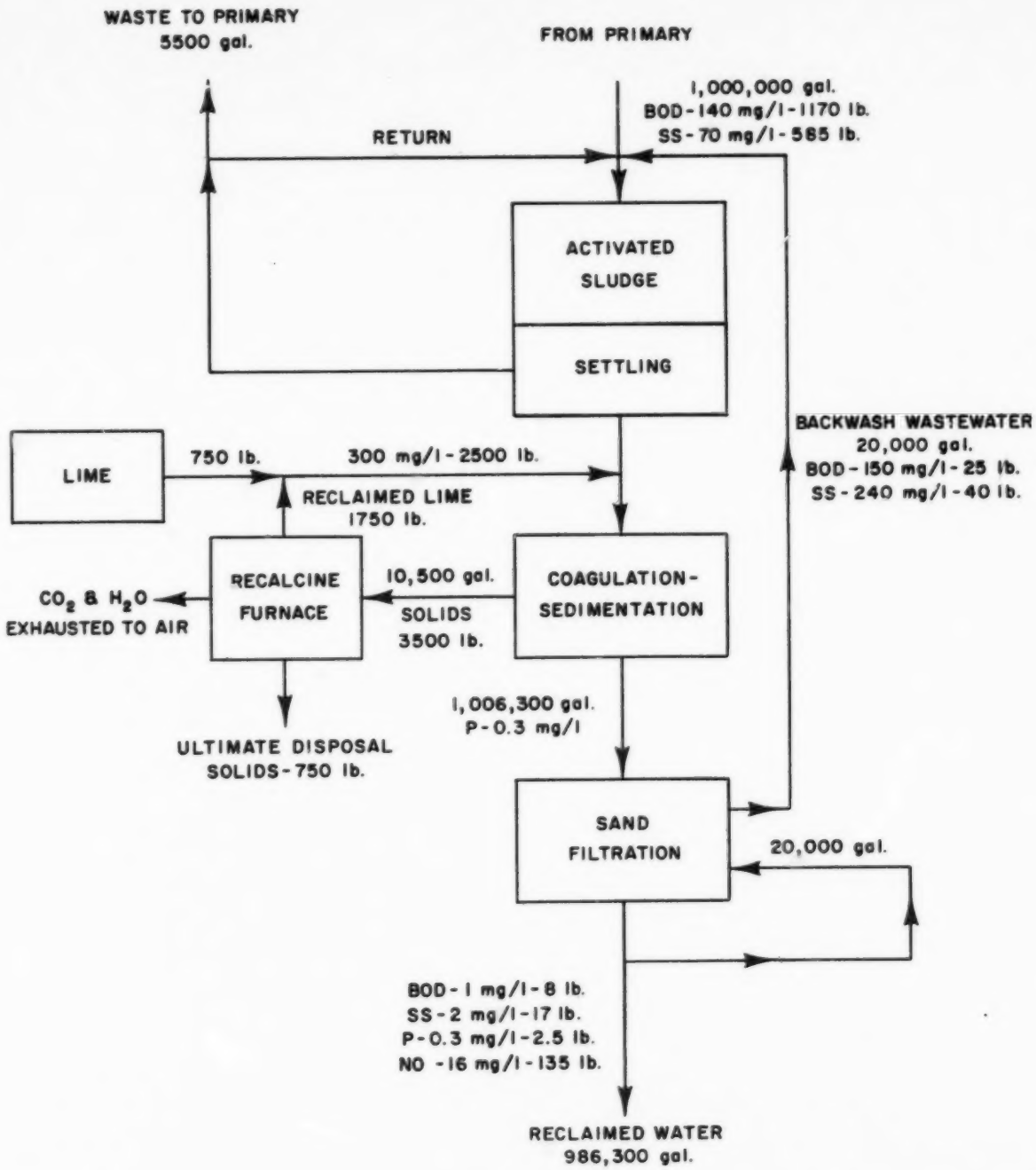


FIGURE 14
WASTEWATER RECLAMATION SYSTEM
PHOSPHATE REMOVAL WITH LIME
MASS BALANCE DIAGRAM - 1 mgd PLANT

result in approximately 5,500 gallons of 11 percent brine and 49,500 gallons of first rinse water containing 450 mg/l chloride water to the effluent water stream would increase the average overall chloride content by 22 mg/l. This should be sufficiently low to permit discharge of the bulk of the rinse wastes to the finished water stream, thereby increasing the total percentage of input water converted to reusable water.

The 5,500 gallons of initial rinse water will be high in chlorides (10,000 mg/l) and may also contain up to 10 mg/l of phosphorus. It could be combined with the waste brine for ultimate disposal. This would involve a total of 11,000 gallons per mg processed (1.1 percent of through-put), having an average brine content of six percent and phosphorus and nitrogen concentrations of 550 mg/l and 1625 mg/l, respectively.

Discussion. In planning and designing facilities to be used for nutrient removal the key factors which must be considered are: 1) the nutrient or nutrients to be removed; 2) the use to be made of the treated wastewater--this will normally govern the required nutrient removal efficiency; 3) the available means for disposing of the ultimate contaminants; and 4) the economic feasibility.

Therefore, once the use to be made of the treated wastewater has been established and the nutrients to be removed have been selected, data similar to that shown on Figures 14 and 15 will have to be developed in order to evaluate the means available for ultimate disposal and the economic feasibility of the process.

Ultimate Disposal of Contaminants--Abstracted from Reference (11).

The quantities of contaminants, both liquid and solid, which will require ultimate disposal will vary the nutrient removal method. For small plants these contaminants can be handled quite readily in lagoons, sludge drying beds, or sanitary landfills. However, as the treatment plant capacity increases, waste tonnages for ultimate disposal may become staggering.

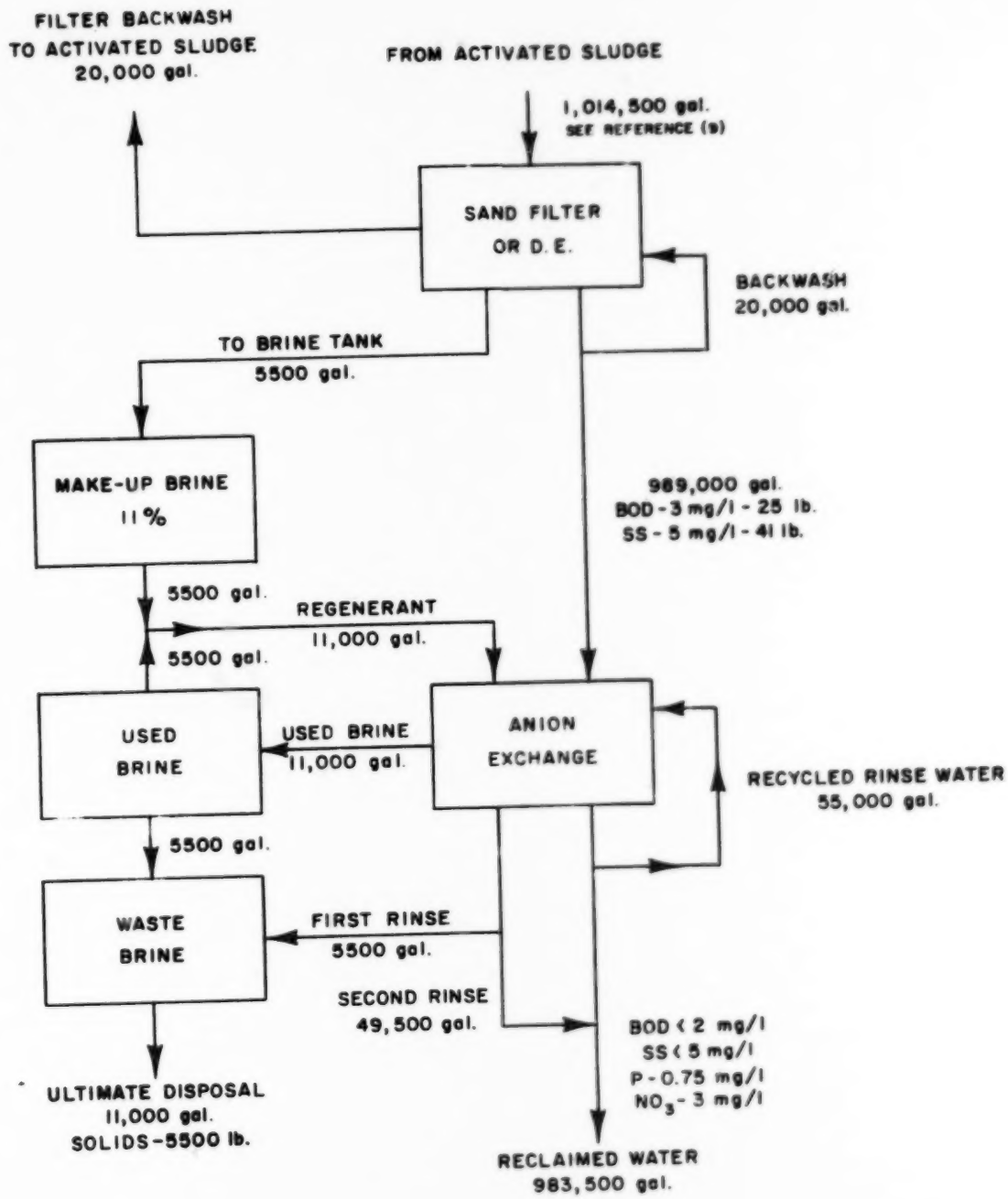


FIGURE 15
WASTEWATER RECLAMATION SYSTEM
ANION EXCHANGE
MASS BALANCE DIAGRAM - 1 mgd PLANT

Various ultimate disposal methods have been studied (12) and a summary of more important methods which are applicable for ultimate disposal of contaminats is given in Table 9. Review of the ultimate disposal methods listed in Table 8 indicates that there are four general possibilities: dumping; subsurface injection; conversion and dumping; and conversion, product recovery, and dumping. Of these four methods the last, although the most desirable, is, at present, the most difficult to obtain. Development of suitable schemes for the recovery of useful products from waste contaminants would go a long way towards bringing about more effective utilization of our water resources.

The cost of ultimate disposal as shown in Table 8 varies over an extremely wide range depending on the disposal method. Due to this wide variation in cost, ultimate disposal methods must be considered carefully since this cost component may be the controlling factor governing the overall feasibility of nutrient removal.

Comparison of Alternative Nitrogen and Phosphorus Removal Methods--Abstracted from Reference (11).

As would be expected, the efficiency, operational characteristics, and cost of the various processes used for nitrogen and phosphorus removal are extremely variable. Nevertheless, comparisons must be made. In order to facilitate the comparison of these various methods, data on the removal efficiency, estimated cost of treatment, and the form of the ultimate wastes to be disposed of have been summarized and are reported in Table 9. Of the information presented in Table 9, the data on removal efficiencies is probably the most reliable since a number of experimental and demonstration studies have been conducted on these processes. The wide spread in the cost data reported for the individual processes reflects the fact that little reliable information is available in the literature.

TABLE 8

ULTIMATE CONTAMINANT DISPOSAL METHODS AND COSTS

Type of Waste	Method of Disposal	Cost[a]	Remarks
Liquid	Soil Spreading Shallow Well Injection	$0.001 - 0.3/1000 gal (8)	Provisions must be made to prevent groundwater contamination.
	Deep Well Injection	$13-27/1000 gal (9)	Disposal site should be available (porous strata, natural or artificial cavities, etc.)
	Landfill	-	Used as a wetting agent.
	Evaporation Pond	$8-64/1000 gal (10)	Provisions must be made to prevent groundwater contamination.
	Ocean Outfall	-[b]	Truck, rail hauling or pipeline needed to transport the wastes.
Sludge	Lagooning		
	Landfill	-	Used as a wetting agent.
	Recovery of Products	-	Depends on waste characteristics.
	Wet Combustion	C.C. $4-80/gpd (8) O and M $1.4-35/1000 gal	Heat value may be recovered for use.
	Incineration	C.C. $10-100/gpd (8) O and M $5-55/1000 gal	Concentration of sludge is needed. Ash must be disposed of.
Ash	Landfill	$0.75-1.0/ton	
	Soil Conditioner	-	Depends on waste characteristics.

Notes:

a. Costs vary with plant capacity and land costs.
b. $25/ft for 8" dia. pipe.

TABLE 9

COMPARISON OF ALTERNATIVE NUTRIENT REMOVAL METHODS

Nutrient Removed	Process Description	Removal Efficiency %	Estimated Removal Cost $/mg	Type of Wastes to be Disposed of	Remarks
Nitrogen	Ammonia Stripping	80-98	8.8-25.0	-	Removal efficiency is based on ammonia nitrogen only.
	Ion Exchange	80-92	170-300	Liquid	Efficiency and cost depend on degree of pretreatment, coag., filt., etc.
	Electrodialysis	30-50[a]	130-210	Liquid	Costs based on 1 mgd-10 mgd capacity solids conc. = 1,000 ppm[b]
	Electrochemical Treatment	80-85	4-8[c]	Liquid and Sludge	
Phosphorus	Chemical Precipitation	88-95	40-70	Sludge	Removal efficiency and cost depend on chemical used.
	Chemical Precipitation with Filtration	95-98	70-90	Liquid and Sludge	
	Sorption	99	39-64	Liquid and Solids	Cost based on water treatment costs.
	Ion Exchange	86-98	170-300	Liquid	
	Electrodialysis	30-50[a]	130-210	Liquid	
	Electrochemical Treatment	81-85	4-8[c]	Liquid and Sludge	

Notes:

a. Removal efficiency per single stage.
b. Cost data obtained from Ionics, Inc., Cambridge, Massachusetts
c. On power cost only; installation costs unavailable.

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Mr. Harold Welborn handled various special investigations and studies relative to the project and assisted in the preparation of this report.

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Professor Eliassen has been with the project since its inception and has spent about ten percent of his time on the project. Biographical data were included in the original application.

Dr. Perry L. McCarty, Professor of Sanitary Engineering, Stanford University.

Professor McCarty has been with the project since its inception and has served as a consultant to the project. Biographical data were included in the original application.

George E. Bennett, Research Engineer, Stanford University.

Mr. Bennett joined the project on August 23, 1965 and spent 100 percent of his time on the project until leaving in April 1967. Biographical data were included in the previous progress report.

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An Engineering Evaluation of Refuse Collection Systems Applicable to the Shore Establishment of the U. S. Navy, Sanitary Engineering Research Laboratory, University of California, Berkeley, California XXV + 478 pp. Feb. 1962, Co-author: G. Klein.

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